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PILOT PLANT EXPERIENCES IN DEMINERALIZATION OF SECONDARY EFFLUENT USING ELECTRODIALYSIS

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PILOT PLANT EXPERIENCES IN DEMINERALIZATION OF SECONDARY EFFLUENT USING ELECTRODIALYSIS*

Carl A. Brunner**

INTRODUCTION

The water resources of the country can be greatly extended if economical ways for treating used water to make it reusable can be found. Some degree of reuse is being practiced now and has been practiced for a long time. With the growth of population, especially in areas of low rainfall, the fraction of unused dilution water must decrease since the total water supply is relatively constant. In some areas of the Southwest the shortage of water is becoming so critical that direct and continual reuse of wastewater will become a necessity. Not only will better removal of organics, solids, and nutrients be required but methods for removing minerals from wastewater that are not now removed by conventional sewage treatment methods must be found.

Municipal secondary effluent contains 300 to 400 mg/l more mineral matter than the water supply to the municipality. Table 1 shows a typical analysis of the secondary effluent from Lebanon, Ohio, where this study was conducted. The tap water to the city is a well water containing about 500 mg/l total dissolved solids. About 350 mg/l solids, therefore, has been added during use. If the water is to be reused many times without

*Presented at the 39th Annual Meeting of the Water Pollution Control Federation, Kansas City, Missouri, September 29, 1966.

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TABLE 1. TYPICAL ANALYSIS OF SECONDARY EFFLUENT, LEBANON, OHIO

Component	Concentration (mg/l)
Total dissolved solids	851
Chlorides	160
Nitrate (as nitrogen)	7.4
Phosphate	25.4
Sulfate	197
Bicarbonate	307
Sodium	123
Potassium	8.2
Calcium	111
Magnesium	30
Ammonia (as nitrogen)	0.2
pH	8.0 units

significant fraction, this mineral increment, about 40 percent of the total, must be removed to prevent the solids content from reaching unacceptable levels. If there is dilution water available the fraction of minerals to be removed would, naturally, be reduced.

Electrodialysis is one method for partially demineralizing water. Bench-scale studies were done under contract to Ionics, Inc. by the Advanced Waste Treatment Research Program of the U. S. Public Health Service (now Federal Water Pollution Control Administration) demonstrated that the process might be economically practical for removing up to one half of the mineral matter from properly pretreated secondary effluent (unpublished data). A pilot plant was constructed at Lebanon, Ohio to determine what operating problems might be encountered in an actual plant, and to realistically estimate the cost of operation. This plant has been operated intermittently for about one year.

DESCRIPTION OF EQUIPMENT

A flow diagram for the electrodialysis stack and pretreatment system is shown in Figure 1. Secondary effluent is fed to a diatomaceous earth filter* for final filtration. The filter aid used originally was a water filtering grade diatomaceous earth.** It was found, however, that better clarity could be obtained if the diatomaceous earth in a 1-lb/gal slurry were treated with 7 percent by weight alum and neutralized with soda ash. The clarified water from the filter enters a storage tank prior to entering

*Bowser-Briggs Filtration Division, Bowser, Inc., Cookeville, Tennessee.
**Celite 503, Johns-Manville Corporation, Manville, New Jersey

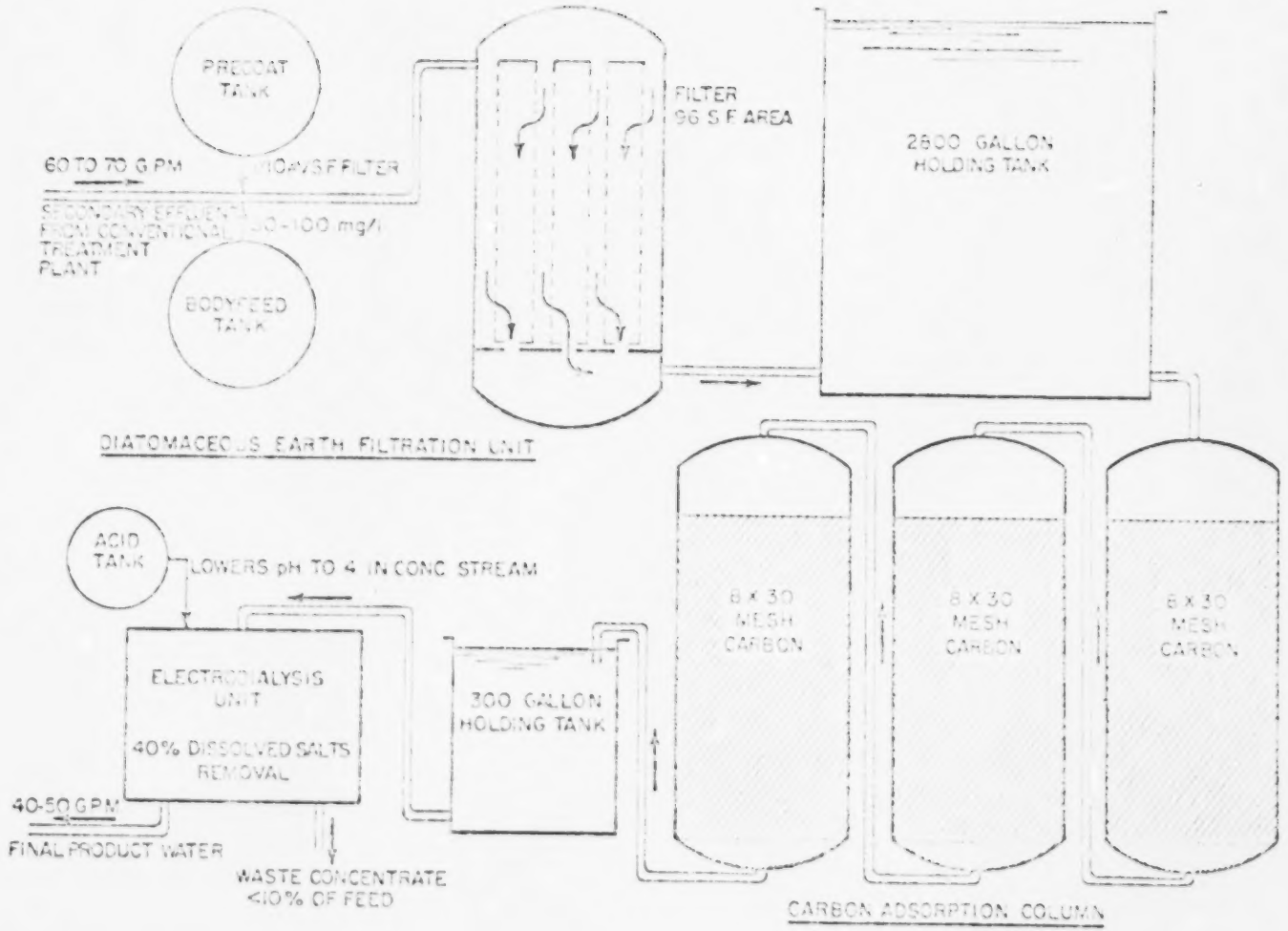

Figure 1. Electrodialysis pilot plant, Lebanon, Ohio.

the activated carbon columns. The purpose of the tank is to provide a supply of water for the remainder of the system during filter backwash. Secondary effluent is used for backwash water.

The water is pumped next through three carbon columns in series for removal of soluble organics. Each column is 4 ft (1.22 m) in diameter, 8 ft (2.44 m) tall and contains 2,580 lb (1170 kg) of 8 x 30 mesh granular activated carbon.* That amount of carbon fills the columns to a depth of almost 8 ft (2.44 m). Across the bottom of each column is a 40-mesh stainless steel screen supported on pea gravel. The steel columns are epoxy lined for corrosion protection. A 6-in. (15 cm) opening at the top is for loading fresh carbon, and a 4-in. (10 cm) opening near the bottom for discharging used carbon. A totalizing water meter is located before the first column.

Water leaving the carbon columns enters a small holding tank which serves as a constant head tank for the electrodialysis stack. Normally a small amount of water is allowed to overflow the tank. The electrodialysis unit is an Ionics Mark III stack which can hold up to 150 pairs of 18 x 40 in. (46 x 102 cm) membranes. It is designed to remove about 40 percent of the mineral matter from the feed water. The membranes are standard brackish water type and dynel backed. They are punched with holes that form internal manifolds for the entering and exiting dilute and concentrate streams when the stack is assembled. Stack assembly and stream flow details are shown in Figure 2. The polyethylene spacers are of the tortuous path type, 0.032 in.

*Type SGL, Pittsburgh Activated Carbon Company, Pittsburgh, Pennsylvania.

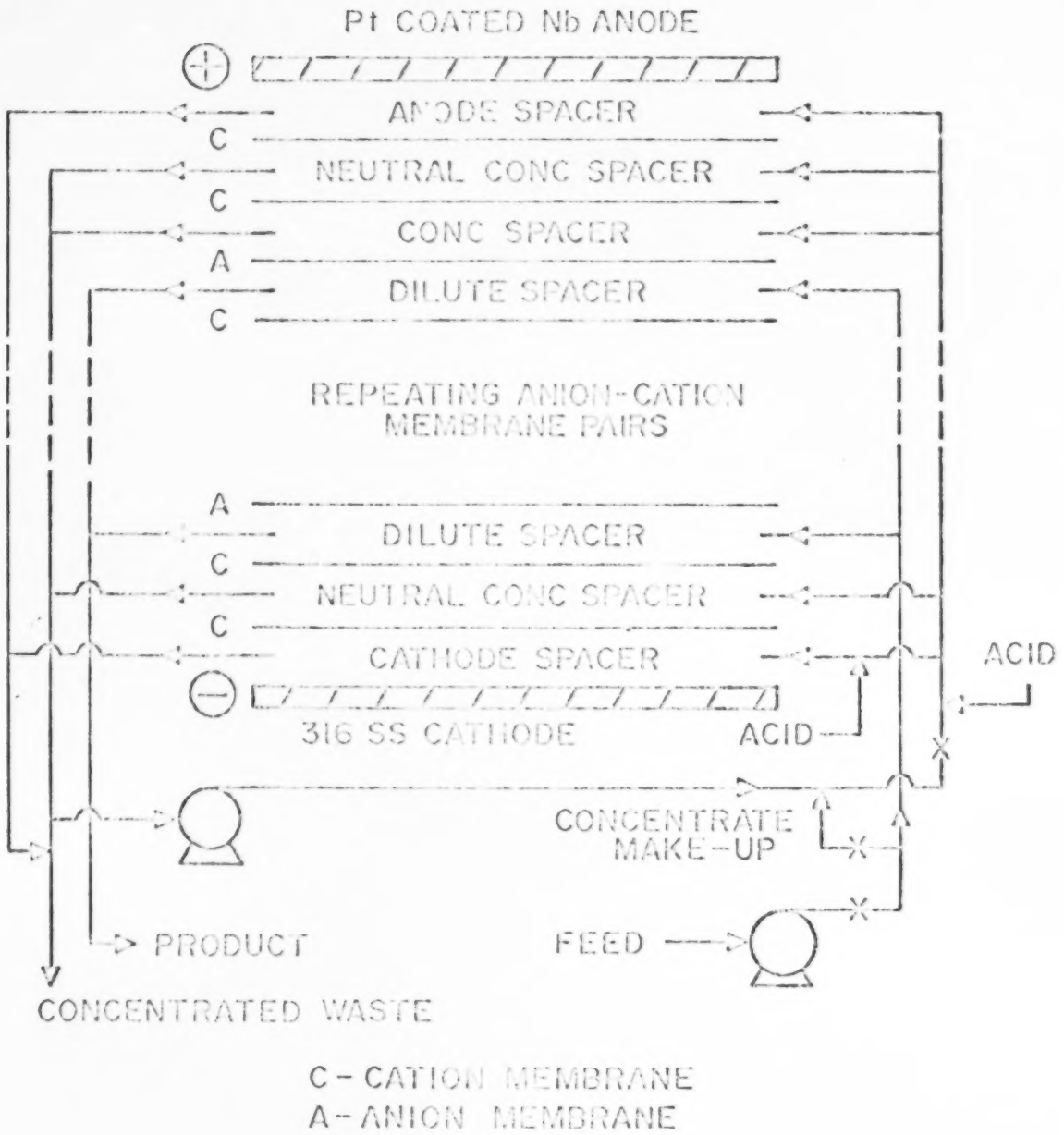

Figure 2. Electrodialysis stack details.

(0.0012 cm) thick, and serve as perimeter gaskets and spacers. The effective open area of the spacers, estimated to be 2960 sq in., is considered to be the effective membrane area. The pressure drop across the stack is about 20 psi (1.41 kg/sq cm) at the design capacity of 1/3 gpm (0.0210 l/sec) per spacer. Electrode spacers are 0.060 in. (0.152 cm) thick. The stack originally contained 150 cell pairs and operated at a product rate of 50 gpm (3.15 l/sec). The number of cell pairs was reduced to 125 and the product rate to 42 gpm (2.65 l/sec), however, to make possible a lower rate through the filter. At 50 gpm (3.15 l/sec) the concentrate makeup was 5 gpm (0.315 l/sec); at 42 gpm (2.65 l/sec) the concentrate makeup is 4 gpm (0.252 l/sec). Electrode compartments each require about 1 gpm (0.0631 l/sec); the rate is not adjustable. Totalizing meters are provided on the concentrate inlet, concentrate outlet, and product line. As can be seen in Figure 2 part of the concentrate stream leaving the stack is recirculated back to the stack after mixing with makeup water. This is done to keep the flow rate through the concentrate compartments nearly the same as through the diluting compartments. This method of operation results in almost equal pressures on both sides of each membrane and prevents membrane and spacer distortion. Sulfuric acid is fed to the concentrate stream and cathode stream using diaphragm pumps. A pH controller is used to control concentrate stream pH in the range of 3 to 5. Cathode stream pH is held at 2 or less. Direct current for the stack is obtained by rectification of a three phase 208-volt source. Capacity of the rectifier is 20 amp. Stack voltage can be varied in approximately 15-volt steps up

to about 165 volts by changing transformer taps. Stack power is measured by a direct current ammeter and voltmeter.

PLANT OPERATION

Operation is continuous for up to 5 days per week. At startup, filtered water is allowed to overflow the large storage tank until it is thoroughly cleaned. Water is then pumped into the carbon columns. After at least 45 min, the approximate time required to displace the water standing in the columns, the carbon column product is pumped to the electrodialysis stack. After flow rates to the stack are adjusted to the desired value the acid pumps and stack power are turned on. Runs are often terminated when filter backwashing difficulties cause the supply of stored clarified water to become exhausted. At the end of a run the carbon columns and electrodialysis stack are kept filled with water.

Carbon column feed and product samples are taken for organic carbon analysis* and turbidity readings** at 4-hr intervals. At 250,000 gal (946,000 l) intervals samples for organic carbon analysis are taken from the bottom of each carbon column to check the degree of exhaustion of carbon in each column.

Electrodialysis feed and product samples are taken at 2-hr intervals, and sometimes more frequently, for conductivity and temperature measurements. From these measurements the approximate degree of demineralization can be calculated. When the samples are taken, flow rates and stack current and voltage are also recorded. If flow rates have drifted from the desired

*Beckman Carbonaceous Analyzer, Beckman Instruments, Inc., Fullerton, California
**Model 1860 Laboratory Turbidimeter, Hach Chemical Company, Ames, Iowa.

values, adjustments are made, but these adjustments are not frequent. The conductivity of the concentrate stream is read at less frequent intervals. Occasional samples of feed, product, and concentrate are taken for determination of total dissolved solids and principal ions.

RESULTS AND DISCUSSION

Membrane Pair

In discussing the results obtained from the operation of the electrodialysis plant, it is useful to consider the membrane pair resistance defined by the equation

$$R = \frac{AN}{nI} - L\rho$$

where R is membrane pair resistance in ohm-sq cm (actually the product of resistance times effective area), A is effective membrane area in sq cm, N is stack voltage, n is number of cell pairs, I is stack current in amperes, L is spacer thickness in cm, and ρ is the mean of the effective specific resistances of the feed and concentrate streams in ohm-cm. The resistance calculated from this equation actually represents more than just the ohmic resistance of the two membranes. Included are membrane potential effects, resistance of the dilute stream-membrane boundary layers and the resistance of any fouling or scaling material that might adhere to the membranes. If it is assumed that the concentrations of the liquid streams in eq/l are directly proportional to the conductances, the current efficiency is the same at every point in the stack, and R is constant across the stack, it can be shown that the effective specific resistance of each stream is the reciprocal of the logarithmic mean of the inlet and outlet conductances

for that stream. Figure 3 shows that the assumption of a direct proportionality between solution conductance and concentration is quite good since the data conform fairly well to a straight line with slope of unity. These conductivity data have been corrected to 25°C by using experimentally determined correction factors. The assumption of constant current efficiency is also probably good. Constancy of R across the stack is greatly in doubt since calculated R values vary considerably with average dilute stream concentration measured here by solution conductance. This variation is shown by the upper plot of Figure 4 where the reciprocal of membrane resistance, or membrane conductance, for data taken at the beginning of a number of experimental runs in a narrow temperature range appears to increase linearly with average dilute stream conductance. The problem of variable feed conductance is unusual in the electrodialytic treatment of water because the ground waters commonly used for feed are of constant composition. The variability must be considered in equipment design, especially when there is significant influx of storm water into the secondary effluent. If a linear variation of membrane conductance with solution concentration is assumed, the calculated membrane resistance must be interpreted as the reciprocal of the log mean membrane conductance over the dilute stream concentration range involved.

Temperature is another variable affecting membrane resistance, or conductance. This is shown in the bottom plot of Figure 4 for runs made within a narrow range of dilute stream average concentrations. This variability also is not always found in electrodialysis applications, but can be important when considering secondary effluent. To describe quantitatively

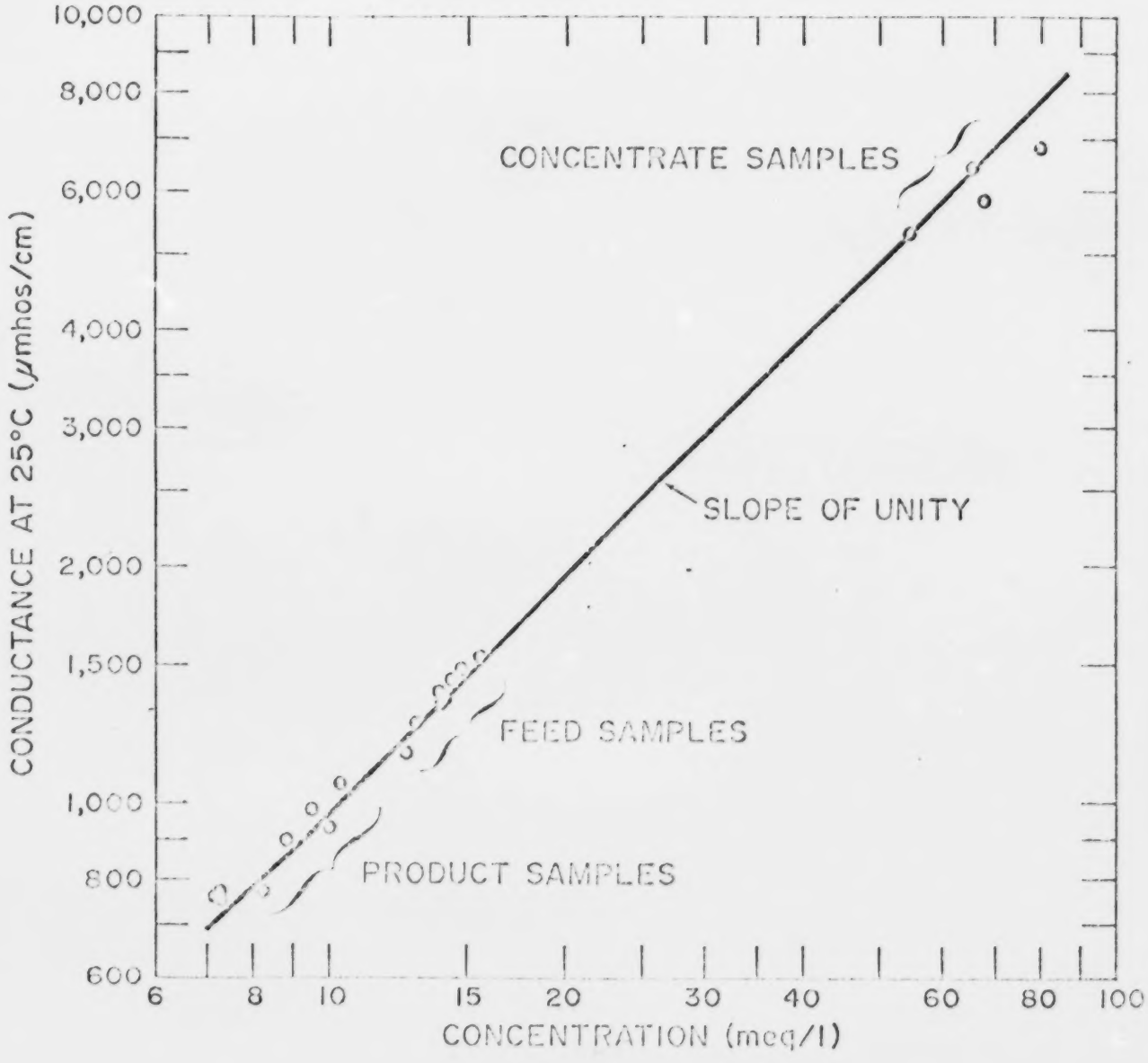

Figure 3. Relationship between solution concentration and conductance.

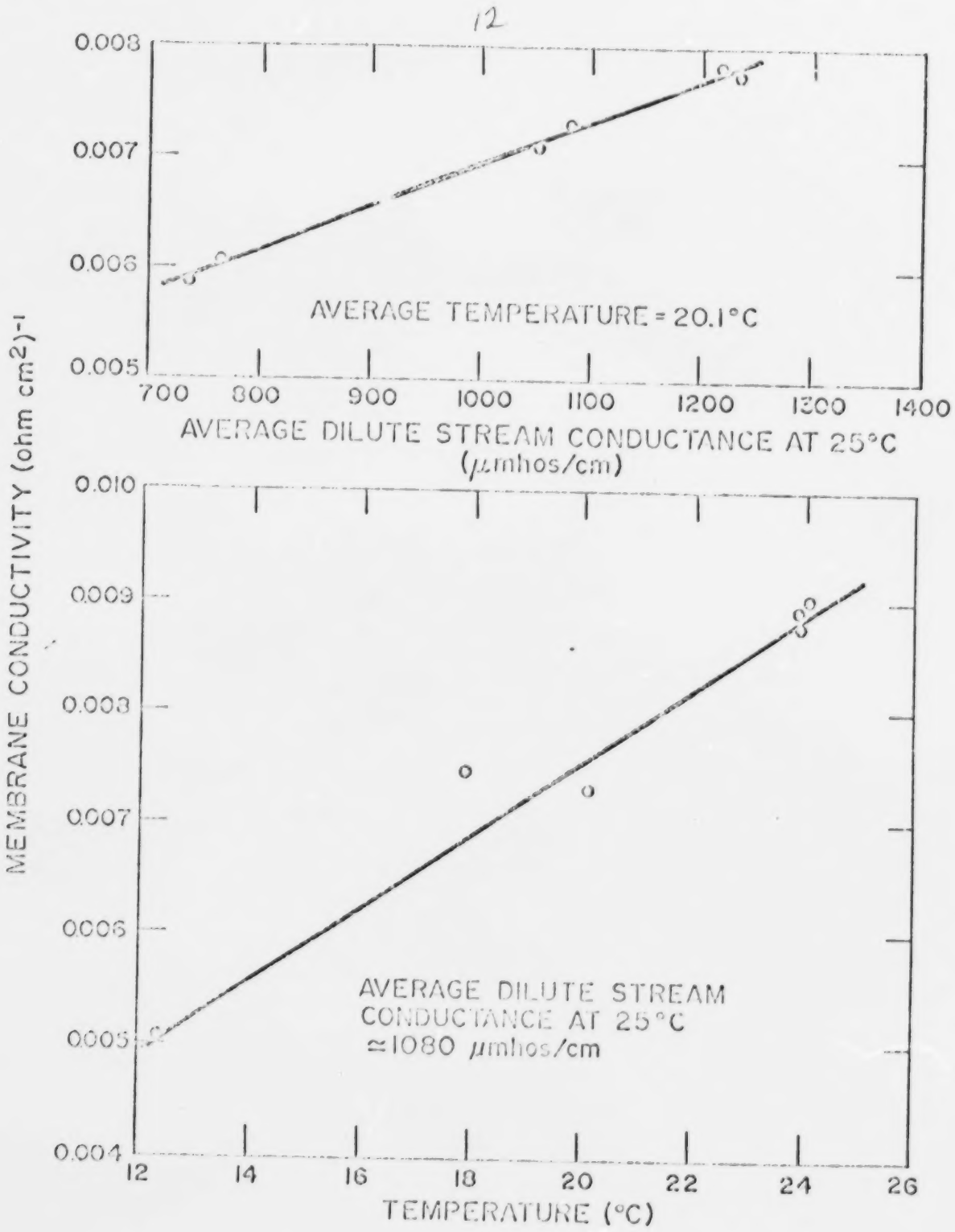

Figure 4. Effect of temperature and dilute stream conductance on membrane conductivity.

the effect of both temperature and dilute stream conductance upon membrane conductivity the slopes of the curves in Figure 4 might be used. To incorporate more data into the determination of these slopes, however, a least squares fit involving the two independent variables was made using 17 R values from data taken near the beginning of all runs up to a total of 351 hr. The range of temperatures covered was 12.0 to 24.1°C, and the range of solution conductances corrected to 25°C was 547 to 1239 μmhos/cm. The reason for using only data at the beginning of a run is to avoid as much as possible membrane fouling effects which might occur as a run proceeded. The slopes obtained were 2.73×10^{-4} (ohm sq cm x °C)$^{-1}$ for temperature effect and 5.01×10^{-6} (ohm sq cm x μmho/cm)$^{-1}$ for solution conductance effect. The intercept value was $- 3.17 \times 10^{-3}$ (ohm sq cm)$^{-1}$. These factors have been found very helpful in correcting membrane resistances to similar conditions for interpreting fouling and other effects.

Membrane Fouling

Membrane fouling was a common difficulty encountered in these tests and has occurred in other operation of electrodialysis equipment. It results in increased membrane resistance and, therefore, decreased stack current and decreased demineralization. During the first four runs totaling 106 hrs there was little fouling. During runs made after that, increasingly serious fouling occurred. The result was a rapid decrease in the fraction of demineralization after startup. Shutting down the stack for a day or more restored it to a reasonable degree of demineralization. The lowest curve in Figure 5 shows the worst case observed. Fraction demineralization

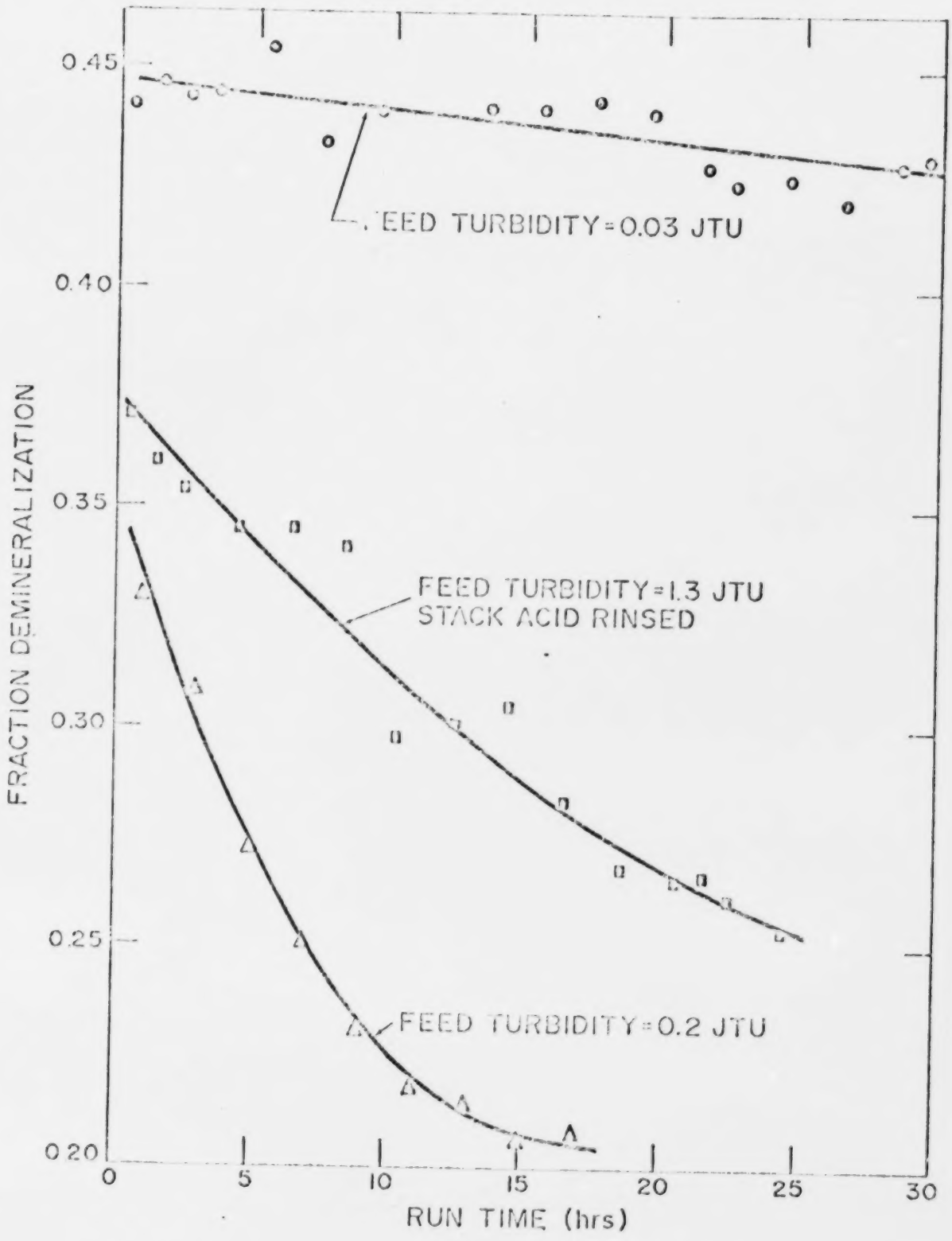

Figure 5. Effect of turbidity on demineralization.

here is calculated from conductivity readings on feed and product. Other runs showed nearly as rapid a fouling rate. Obviously operation of an electrodialysis stack under these badly fouling conditions would not be practical.

Considering the source of feed and the type of pretreatment used, the suspected cause of stack resistance increase would be accumulation of colloidal and possibly trace soluble organics upon the dilute stream side of the anion membranes. These membranes rather than the cation should be affected because of the negative charge on the organic material. In the applied electric field this material would tend to move toward the anion membranes. Examination after the stack was badly fouled showed that the dilute stream side of the anion membranes had collected a thin coating of slimy material.

Because the fouling material built up as a fairly easily removable layer on the membrane surface rather than penetrating the membrane, colloidal material would be the likely cause of the trouble. Berger and Chew[1] showed that soluble organics similar in nature to those likely to be found in the feed to the pilot stack penetrate the membranes and are very difficult to remove. The activated carbon treatment gave a feed with organic carbon content averaging 2 mg/l or less for all runs. This treatment appears adequate for preventing serious fouling by soluble organics.

If colloidal material were the cause of fouling, the fouling rate should be related to the turbidity of the feed or more specifically the difference between feed and product turbidity. Spot checks of product turbidity have not consistently shown a smaller value than in the feed.

The f[illegible] material actually depositing on the membranes could l[illegible] that the effect on turbidity could be masked by the lo[illegible] measurements. That the magnitude of the feed turbidity [illegible] fouling rate can be seen by comparing the top and bot[illegible] Figure 5. These curves are typical of results of runs made u[illegible] of roughly the same turbidities as given on the figure[illegible] that the wide variation in turbidity is not a result o[illegible] method of clarification. For some unknown reason the dia[illegible] produced lower turbidity water as the temperature increased.

Th[illegible] relation between turbidity and fouling rate. All ru[illegible] between 0.2 and 0.8 JTU resulted in less rapid f[illegible] own with a turbidity of 0.2 JTU. This lack of a co[illegible] ly demonstrated by the middle curve in Figure 5 [illegible] from a run where the feed turbidity was unusual[illegible] ing this run the stack was rinsed with water acidifi[illegible] a disinfectant. The unexpectedly low results [illegible] to that growth of microorganisms on the membra[illegible] sistance or increases the thickness of the deposit[illegible] ple examination of the fouling material has always [illegible] large numbers of bacteria, protozoa, and higher [illegible] of these organisms deposits from the water and wa[illegible] grow on the membranes is difficult to determine[illegible] counts on two feed and two product samples

taken when turbidity was very low indicated that some bacteria are removed during passage through the stack. More study of this aspect of membrane fouling is required.

Table 2 summarizes results of runs made using low turbidity feed water. The periodic addition of iodine is indicated. The effect of the iodine on fouling is in most cases inconclusive. Differences in demineralization values at the beginning and end of a run indicate the effect of fouling. With turbidities in the range shown runs of several days are possible before the demineralization falls below an adequate value. By moderately increasing the voltage as fouling occurs it should be possible to extend the practical operating time to 7 days or more. At that time some method for removing fouling material would have to be employed. Shutting the stack down for a day should suffice.

Diatomaceous earth filtration as used in this application did not consistently produce water of very low turbidity. A chemical clarification system is expected to consistently produce water of the required clarity and is included in plans for further work.

Fouling has been discussed in terms of the rate of decline in demineralization. A more basic parameter for describing fouling, and an obvious one to use if a correlation can be found in terms of feed properties, is rate of change of membrane pair resistance. Figure 6 demonstrates an instance in which consideration of membrane resistance is necessary to properly describe the results of operation. Variation of feed concentration, temperature, and line voltage during this run resulted in a demineralization curve that suggests a much higher fouling rate during the first 13 hr segment of operation than immediately after. The membrane resistance-time curves

TABLE 2. LOW TURBIDITY ELECTRODIALYSIS RESULTS

Average Feed Turbidity (JTU)	Run Length (hr)	Starting Fractional Demineralization	Final Fractional Demineralization	Comments
0.004	76.5	0.43	0.37	Cleaned stack with I_2 shortly before starting
0.009	33.0	0.43	0.37	
0.025	89.3	0.44	0.41	Added 0.6 mg/l I_2 for 6 hr near middle of run
0.021	37.5	0.44	0.41	
0.12	51.9	0.43	0.39	Added 1.2 mg/l I_2 for 1 hr near start of run
0.010	36.0	0.46	0.44	Added 1.4 mg/l I_2 near end of run
0.015	29.5	0.45	0.43	Started 7.75 hr after end of last run
0.019	20.4	0.45	0.43	
0.007	18.8	0.45	0.43	
0.010	88.6	0.44	0.37	

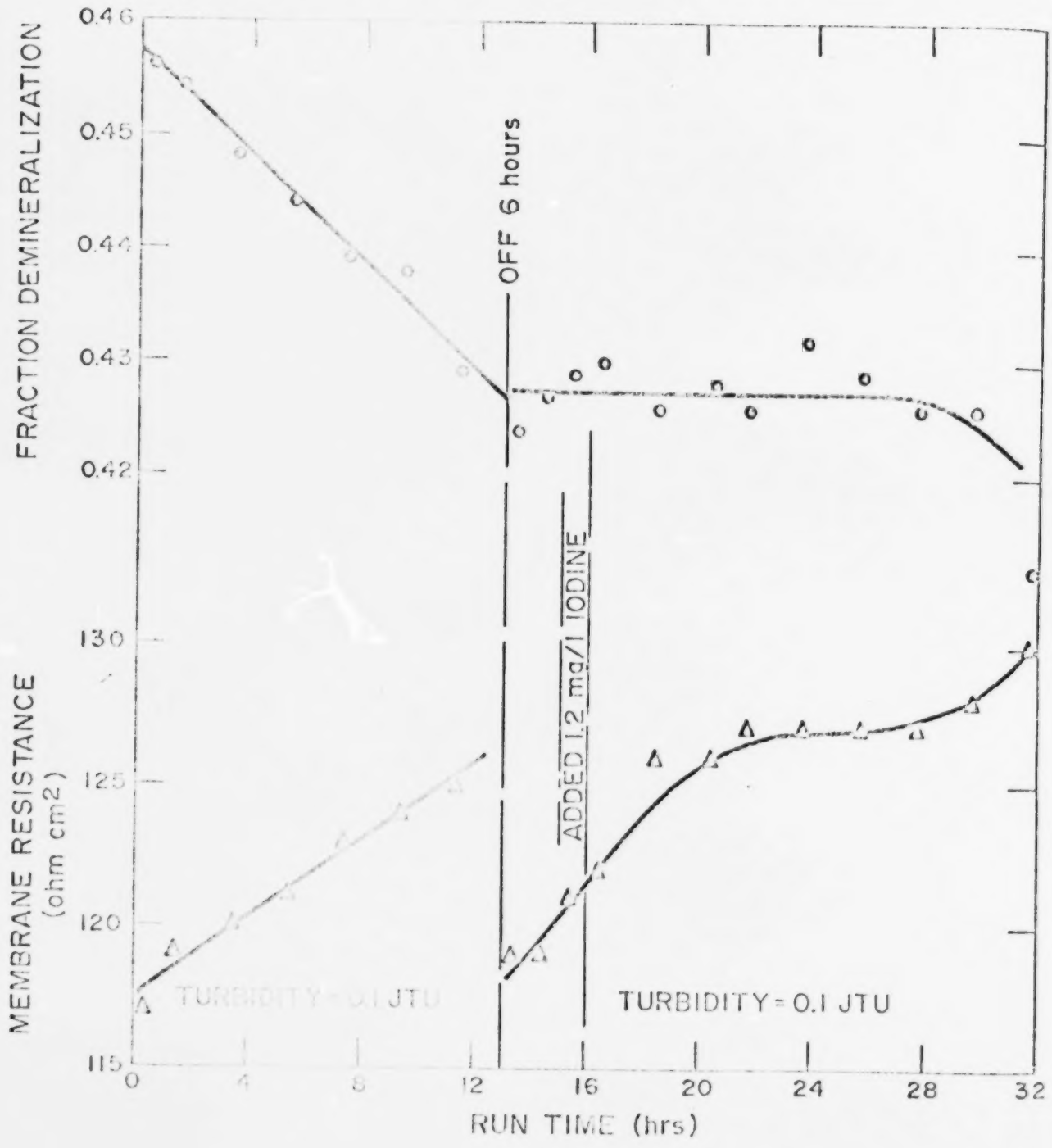

Figure 6. Effect of variables beside membrane fouling on change in fraction demineralization.

show that the rate of fouling was actually slightly greater at the beginning of the second segment of the run than for the first segment.

The use of iodine in the feed water as indicated in Figure 6 was an attempt to decrease the rate of fouling by killing microorganisms on the membranes. The fact that stack fouling was not obvious just after installation and before the membranes had become contaminated suggested the possible importance of growth of organisms. Later, use of acid as a disinfectant at the beginning and during runs appeared to decrease fouling but not enough to make operation practical at turbidities above 0.1 JTU. Silver nitrate was next investigated, but showed no beneficial effect. By the time iodine was tried feed turbidities had dropped to very low levels so that fouling was slow. The leveling off of the membrane resistance curve in Figure 6 a few hours after feeding iodine is a possible indication of the effect of disinfection. Resistance curves for other runs using iodine as indicated in Table 2 did not give this result. The slopes of the resistance curves, however, were less than those in Figure 6. Control of turbidity, possibly combined with a periodic disinfection, presently appears to be a more feasible method for keeping fouling within practical limits than dependence on disinfection.

Long Term Change in Membrane Resistance

Although fouling has been severe at times, it has not been permanent for the most part. Better pretreatment will likely eliminate it as an important factor. Shutting down the stack for a day or more, restores the demineralization to a reasonable value. There could be, however, slight permanent damage to the membranes from fouling and from causes independent

of fouling. Figure 7 gives a history of membrane resistance from data taken at the beginning of runs. The data have all been corrected to a standard set of conditions for realistic comparison. The corrections were based on data for the ft at 351 hr of operation as discussed earlier. The upper plot shows how these data scatter about the value of 130 ohm sq cm calculated from the least squares equation for membrane conductance as a function of temperature and solution conductance. The correction factors are admittedly not completely satisfactory, but offer the best method available for putting data on a comparable basis. Since data usually were not taken less than 0.5 hr after a run was started, and since fouling was sometimes rapid enough to change the membrane resistance slightly in that time, resistance values extrapolated to zero run time were used in some cases.

Membrane resistance appears to increase significantly after about 400 hr. This was a period of very severe fouling and the high resistance values probably result from the thin layer of slime that remained on the membranes even after periods of shut down.

From about 500 to 800 hr various acid treatments and silver treatment were investigated. Because of uncertainties in solution conductivities at low pH, membrane resistances for some runs were not calculated during this time. It was during this period that calcium carbonate scale was first found on the concentrate side of the anion membranes. This scale not only formed on the surface, but became imbedded in the membranes. Resistances since then probably show some effect from the scale which permanently damages the membranes even though it is removed.

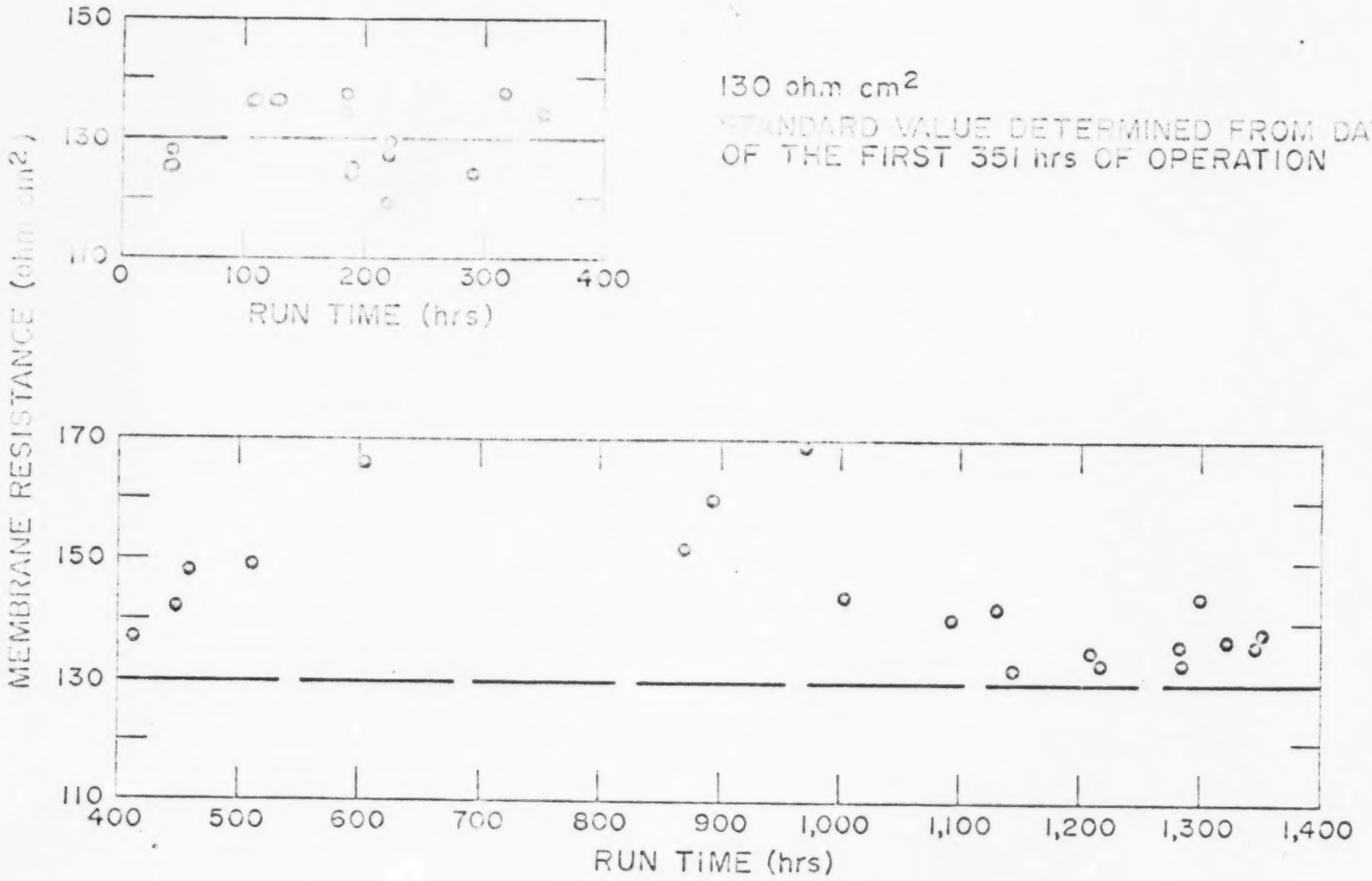

Figure 7. Running record of starting membrane resistance corrected to 20° C and 1100 μmhos/cm dilute stream conductance.

Feed turbidity was nearly always less than 0.1 JTU and sometimes about 0.01 JTU after 890 hr. This was a period of low fouling. Except for the point at 972 hr, membrane resistance tended to decline to a range not greatly different from the average found when the stack was new. The reason for the unusually high data point is not clear. Interestingly this run showed a higher fouling rate than other runs in the same feed turbidity range. The membranes were cleaned with tap water at the end of the run. Membrane resistance declined sharply.

The most important result demonstrated by Figure 7 is that membrane resistance did not show a large permanent resistance increase after severe fouling that occurred during more than 500 hr of operating time. The increase of about 5 percent that did occur could quite reasonably be attributed to scale damage and adsorption by the membranes of traces of soluble organics from the water. Adsorbed material might be partially removable by chemical treatment. No treatment has been attempted, however. Assuming that fouling can be kept to levels found during the last 500 hr of operation, further increase in resistance might well be expected to be at a lower rate than in the past. A much longer period of operation will be necessary to fully evaluate membrane life.

Membrane Scaling

If the volume of concentrated waste from an electrodialysis stack were not an important factor, it would be possible to operate so that there was no problem of scale formation because the mineral content is low enough that scaling materials, most likely calcium carbonate and calcium sulfate, are far from their maximum solubility. On the other hand it

would probably be desirable in most cases to operate with as little waste concentrate as possible. Scale formation then becomes the limiting factor in determining the volume of concentrate. In most of this work, the waste has been kept at 9.5 percent of the product volume. This gave a maximum solids content in the concentrate stream of about 6000 mg/l, a maximum calcium content of about 700 mg/l and a maximum sulfate content of about 2000 mg/l. In view of the fouling problems encountered it was not felt desirable to purposely introduce scaling problems. Some acid soluble scale did form, however, during some runs. Acidification of the scale showed it to be largely calcium carbonate. Further analysis was not made. This scale occurred even though the concentrate stream was acidified to pH values of 3 to 5 where carbonate ion concentration becomes negligible. Since the membrane surface can be at pH conditions considerably different from the bulk solution, it is not surprising that scale formed. Lower pH in the concentrate stream could be used. It is doubtful, however, that the volume of concentrate waste could be safely reduced much below the 9.5 percent level for the particular water studied. It must be remembered that there is variability in wastewater composition over a period of a day or even hours. Operation too near the point of scale formation should be avoided. For feed water lower in calcium than used here the concentrate stream could be reduced in volume, possibly very significantly.

Ion Removal Selectivity

In the process of electrodialysis not all ions are necessarily removed to the same degree. This depends upon the particular type of

membrane being used in the solution composition. Table 3 gives ion selectivity data from eight sets of feed and product samples taken since operation was begun. In the table the fraction of all ions removed is based on the total normality of feed and product samples assuming no ions other than those shown are present. Probably no other ions are present in amounts great enough to significantly affect the results. The membranes are not highly selective for any particular ion. Highest selectivities are for calcium and chloride; the lowest are for sodium and phosphate.

Although there is considerable scatter in the data for most of the ions, phosphate and sulfate have shown definite trends since operation began. Phosphate has shown a decline in selectivity from 1.03 to 0.25. The reason is not clear but could be related to membrane fouling. Sulfate showed a large decline over the first 370 hr of operation from a selectivity of 2.0 to 0.21. This decline was probably associated with the formation of a fouling layer on the anion membranes. Later data, although scattered, again showed increased selectivity but not to the very high value shown when the membranes were new.

Current Efficiency

Current efficiency is important from an economic aspect and because it is an indication of undesirable polarization at the membranes and loss of permselectivity of the membranes. Polarization occurs when there is excessive splitting up of water into ions and can lead to scale formation and membrane damage. The transport of the water ions, especially hydroxyl ion, from the dilute stream to the concentrate stream lowers the efficiency for removal of mineral ions. Permselectivity is a measure of the ability of a membrane to transmit cations but not anions and vice versa.

TABLE 3. ION REMOVAL SELECTIVITY

Ion	Fraction of Ion Removed / Fraction of All Ions Removed	
	Average	Range
Bicarbonate	0.77	0.63 - 0.95
Chloride	1.25	1.07 - 1.51
Nitrate	1.22	1.02 - 1.48
Phosphate	0.72	0.25 - 1.03
Sulfate	1.11	0.21 - 2.00
Ammonium	1.15	1.05 - 1.3
Calcium	1.25	1.01 - 1.48
Magnesium	0.93	0.71 - 1.35
Potassium	1.14	0.77 - 1.35
Sodium	0.79	0.52 - 0.99
Calcium and Magnesium	1.13	1.05 - 1.28

Considering leakage of current through interval manifolds in the stack, the maximum fraction of the current available for demineralization is slightly less than 0.95. The average value found from runs where the total ion concentration of feed and product were known was 0.86 with a range from 0.79 to 0.92. Two low values, 0.79 and 0.83 were found to occur when the membranes were badly fouled and could indicate that polarization was taking place. The average without the two low values was 0.89. There has not been a noticeable decline in the current efficiency indicating that permselectivity has not changed significantly.

TREATMENT COST

Although the problem of membrane fouling has not been completely solved, it is felt that the low fouling rate obtained with low turbidity feed water makes operation practical. Periodic shutdowns to allow the stacks to recover capacity probably would be necessary. Periodic current reversal or use of a fouling scavenger treatment before the stack might also be practical.

The following capital and operating cost estimates are made assuming low turbidity feed water and equipment of the same type as used in the pilot plant. A 15 percent excess in electrodialysis stacks is included to take care of shutdowns associated with fouling problems. The production rate assumed is 10 mgd (37,850 cu m/day) for 365 days per yr. Water of composition similar to the pilot plant feed and with a maximum total dissolved solids of 650 mg/l will be demineralized to 500 mg/l. The conservative 1/2 gpm (0.0210 l/sec) per cell pair used in the pilot plant

will be assumed. For 275 cell pairs per stack there will be 83 stacks required. The maximum rectified power will be about 370 kwh.

Table 4 gives the details of the capital cost for the plant. Stack costs are estimated from prices of replacement parts.

Table 5 gives the operating costs for the plant not including pretreatment or concentrate disposal. The estimated value of 16.0¢/1000 gal (4.21¢/cu m) is considered conservatively high. It is quite likely that the rate of 1/3 gpm (0.0210 l/sec) per cell pair can be increased considerably without causing serious polarization problems. If this rate were increased to 1/2 gpm (0.0316 l/sec) per cell pair, for example, the number of stacks could be reduced to 59 and the capital cost would be reduced to $2,314,000. This would reduce the operating cost to 11.9¢/1000 gal (3.14¢/cu m). Considerable further reduction in capital cost would result if the costs of membranes and spacers were reduced. Until the market for these increases greatly, however, significant reductions are not likely. The present demand does not permit manufacturers to invest in the continuous and more highly automated equipment necessary for producing cheap stack components.

SUMMARY AND CONCLUSIONS

A pilot scale electrodialysis stack has been operated on diatomaceous earth filtered and activated carbon treated municipal secondary effluent. Soon after operation was begun serious membrane fouling occurred. The fouling materials were quite effectively removed by shutting down the stack for periods of a day or more. The fouling rate is related to feed turbidity, but a precise correlation does not exist. There is evidence

TABLE 4. CAPITAL COST FOR A 10 MGD ELECTRODIALYSIS PLANT

Item	Cost ($)
Electrodialysis stacks	
Membranes	8,370 per stack
Spacers	8,290
End plates	120
Electrodes	815
Other parts	100
	17,695 per stack
	1,557,200 for 88 stacks
Installation - 15% of component cost	233,600
	1,790,800
Rectifiers - installed	43,900
Pumps and motors - installed	
10 feed and 10 concentrate	45,700
Acid - 3 per stack	52,800
	98,500
Piping - 30% of installed cost of stacks and pumps	566,800
Instrumentation and controls - 10% of stacks and pumps	188,900
PVC lined acid storage tank - 25,400 gal (96.1 cu m) - installed	11,000
Building - 100 sq ft/stack at $13/sq ft	114,300
Contingencies - 10% of investment	281,400
Engineering - 5% of above	154,800

TABLE 4. CAPITAL COST FOR A 10 MGD ELECTRODIALYSIS PLANT (Continued)

Item	Cost ($)
Interest on investment during construction - 4% of above	130,000
Total capital cost	3,380,400

Note: mgd x 3785 = cu m/day
gal x 3.785×10^{-3} = cu m
sq ft x 9.29×10^{-2} = sq m

TABLE 5. OPERATING COST FOR A 10 MGD ELECTRODIALYSIS PLANT

Item	Cost (¢/1000 gal)
Electric power	
Maximum rectifier power (1 0 kw at 90% rectifier efficiency and power at 0.7¢/kwh)	0.69
Pumping power (586 hp at 0.7¢/kwh)	0.32
Other power (negligible)	--
	1.01
Sulfuric acid (to neutralize 189 mg/l HCO_3^- and at 1.3¢/lb)	1.56
Maintenance	
Membranes (4 sets over 25 years)	3.23
Spacers (1 set over 10 years)	2.00
Anodes ($200 per stack per year)	0.48
Other materials (0.5% of remaining capital cost per year)	0.25
Labor (0.5% of capital cost per year)	0.46
	6.42
Operating labor (3 operators at $6000/year and one operator and coordinator at $8000/year)	0.71
General and administrative overhead (30% of labor cost)	0.35
Amortization of equipment (25 years at 4% interest)	5.92
Total operating cost	15.97

Note: mgd x 3785 = cu m/day
¢/1000 gal x 0.264 = ¢/cu m
lb x 0.454 = kg

that growth of microorganisms on the membranes could also affect the fouling rate. Permanent damage to the membranes from fouling has been slight with an increase in membrane resistance of only about 5 percent during more than 1,400 hr of operation, much of which was under severe fouling conditions. Current efficiency has not declined noticeably during this time.

Operating with concentrated waste of about 9.5 percent of the product has resulted in periodic minor scaling by calcium carbonate. This is not considered a serious problem.

Assuming feed turbidity of less than 0.1 JTU can be maintained, the membrane fouling rate is low enough that operation of electrodialysis becomes practical without the use of auxiliary means for fouling prevention. Using operating conditions similar to those in the pilot stack the estimated total cost of operation is 16.0¢/1000 gal (4.2¢/cu m) for a 10 mgd (37,850 cu m/day) plant. By increasing the rate through the stacks it should be possible to reduce the cost to 11.9¢/1000 gal (3.1¢/cu m).

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